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INVESTIGATIONS ON PEAT ENERGY FOR PROCESS HEAT AND MANGANESE PRODUCTION

DSS File No.: 51SQ 23216-6-6286

FINAL REPORT

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Report No.: AS/87/27 (Revised)
Job No.: 05000234
Date: August 19, 1987

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89-05735/

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SUMMARY

This is the final report for DSS Contract No.: 23216-6-6286 entitled "Investigations on Peat Energy for Process Heat and Manganese Production". It was funded under the 1986/87 Bio-energy Development Program of EMR to assess the possibility of energy generators from peat using circulating fluidized bed combustion technology. Test work was carried out on both thermal and manganese peat.

The combustor was operated reliably with both peat types at varying excess air levels. The thermal peat achieved a high conversion of carbon to gaseous products (combustion efficiency of 97.8%) operating at an excess air level of 13.5% and 40% moisture in the feed. Two tests with manganese peat were performed at reducing conditions of -15.6% and -35.6% excess air levels. High carbon conversions, (low carbon losses in the ashes) were obtained 93.7% and 94.7% for tests 1 and 2, respectively.

Reducing conditions for the manganese peat test were selected with a view to subsequent recovery of the metal form the ash which could be an interesting proposition. However, this was not part of this project.

1.0 INTRODUCTION

New Brunswick has extensive peat resources (103 million tonnes occupying an area of about 60,000 hectares) which at present are primarily utilized for horticultural purposes. Two types of unutilized peat are thermal and manganese.

The thermal peat is similar in composition to normal horticultural peat but has a higher degree of humification (decomposition) which makes it unsuitable for soil conditioning. This material has a heating value of approximately 4000 to 5000 BTU per pound (50% moisture content) and an extremely low ash (1%) and sulphur content, which make it an ideal fuel.

It is estimated that the thermal peat resources in New Brunswick are approximately 21 million tonnes. Furthermore there could be problems associated with the use of thermal peat as a fuel source on a large scale (1). Therefore it's utilization in New Brunswick would likely be on a small scale for local users as it is the case for wood energy in the province.

The managanese peat is very different in composition to thermal and horticultural peat. Unlike thermal peat it has a much greater ash content which as its name suggests, contains large amounts of manganese. The calorific value of the manganese peat is lower than thermal peat because of this high ash content. The manganese peat used in this study had a calorific value of 2826 Btu/lb (22% moisture content). Therefore, using manganese peat as a fuel source is less attractive but if the manganese metal is extracted from the ash resulting from the combustion of the peat, the situation could change. Manganese is a metal used principally in iron and steel manufacturing and since there is no domestic supply, Canadian production would be attractive. Annual consumption of manganese in Canada is in the order of 250,000 tonnes of ore and 4000 tonnes battery and chemical grade manganese.

The primary objective of this study was to combust thermal and manganese peats in the circulating fluidized bed combustor

(CFBC) at the Research and Productivity Council and to evaluate the performance of both as fuels.

A circulating fluidized bed combustor (CFBC) can be considered the logical extension of bubbling fluidized bed combustors (BFBC). The main difference in operation between the two units is the gas velocity used. The CFBC operates at much higher gas velocities and as a consequence has a much different design configuration and differing combustion characteristics. A comparison of the two systems is shown in Figure 1.

A BFBC operates at low gas velocities to maintain a distinct dense bed of solids and to prevent entrainment of fine solids in the gas stream. This results in a low relative velocity between the gas and the solids (slip velocity) and can lead to poor mixing of the solids and gas. This allows combustion air to bypass through the bed without reacting. As a consequence the combustion reactions in a BFBC are controlled by the rate of mass and heat transfer in the bed.

A CFBC operates at much higher velocities than a BFBC. These velocities are well above the entrainment velocity of an individual particle. To maintain a dense bed of solids in the reactor and to prevent the combustor from operating as a transport reactor requires the recirculation of large amounts of solids back into the reactor. This causes clustering of particles which behave like large particles and reflux within the combustor. This results in a high particle slip velocity and excellent heat and mass transfer. Therefore, combustion reactions are controlled by the kinetic limitations. The high degree of mixing obtained by circulating the solids also simplifies the feeding of fuel into the system since fewer injection ports are needed as compared to a BFBC.

A CFBC has the same advantages as a BFBC in terms of long particle retention times allowing low operating temperatures in the order of 850°C. This results in low NO_x emissions which can be further reduced by splitting the air flow into primary air at

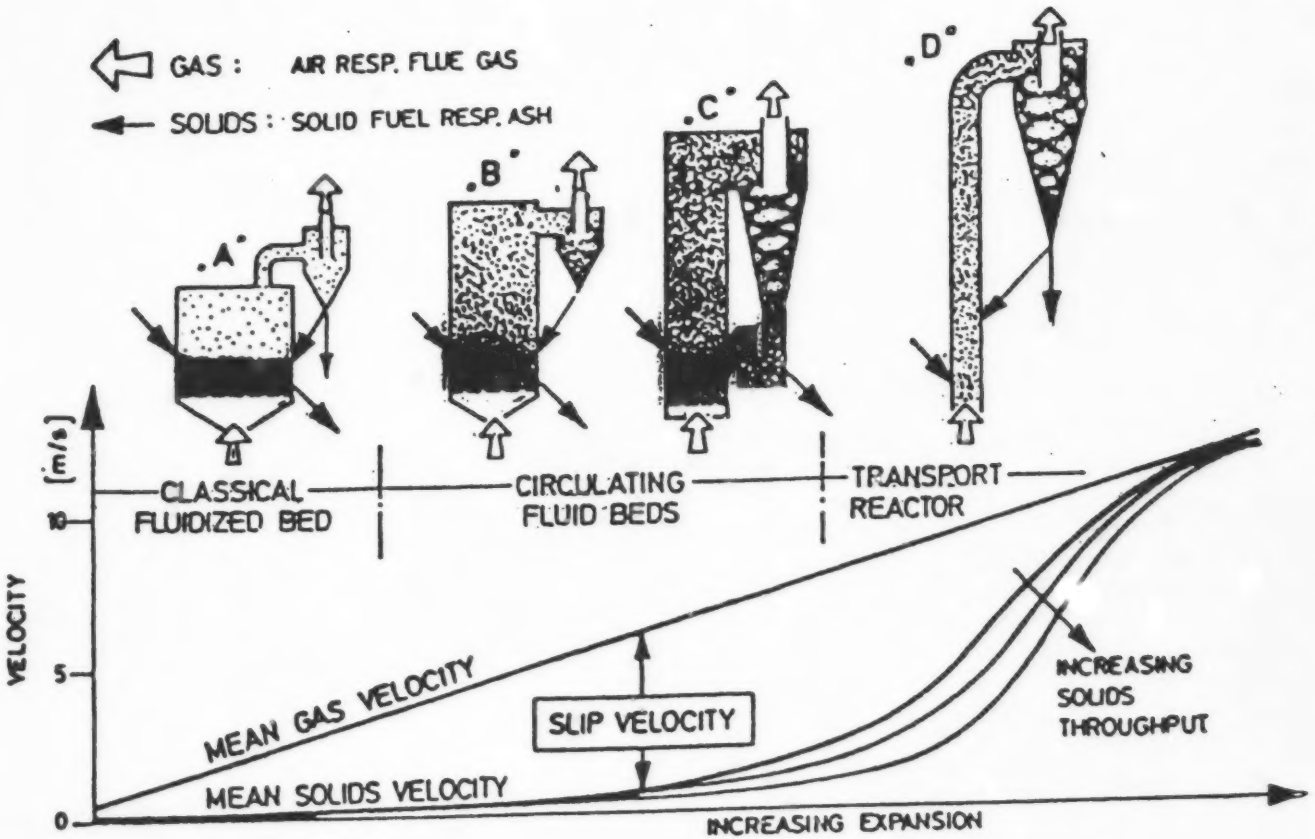

Figure 1: Basic Systems of Gas/Solid Reactors

the base of the unit and secondary air at a location higher in the column of combustor. By this staging of the air a reducing zone is created at the bottom of the combustor which discourages NO_x formation. Staging (i.e. high secondary air) can be used without danger of defluidizing and slumping of the bed.

The wide range of operating velocities in a CFBC also allows for better control as compared to a BFBC. The thermal output of the combustor can be easily reduced by reducing the operating velocity and fuel. Also rapid load following is possible because of the large reservoir of heat contained within its inventory of solids. The CFBC is also capable of operating with a range of fuels because of its excellent mass and heat transfer between the gas solids.

The heat extraction surfaces located in the CFBC combustor are located along the walls as opposed to in bed tubes. This helps to prevent tube erosion which can occur in a BFBC. It is also possible to vary the heat removed by the "water wall" cooling surfaces in the CFBC by changing the recycle rate of solids into the reactor. This will alter the solids loading, which in turn will vary the heat transfer coefficient of the water wall.

The contract was awarded under the 1986/87 Bio-energy Development Program of the Renewable Energy Division, Energy Mines and Resources. A copy of the work statement is given in Appendix A. It meets the objectives of the program dealing with feedstock preparation proposals and fuel conversion technology with the aim of advancing the exploitation of peat fuels.

The thermal peat was combusted in normal oxidizing conditions. The manganese peat was combusted in reducing conditions. This was necessary so that the manganese was converted into the monovalent form which is required in further treatment of the ash to recover the manganese.

2.0 EXPERIMENTAL SET-UP

The circulating fluidized bed combustor system is shown schematically in Figure 2. It consists of a combustion chamber, a primary cyclone, a stand pipe, a heat exchanger, a secondary cyclone and a baghouse. Solids are fed to the system pneumatically. A brief description of the major components is given below.

The Combustor

The combustor is cylindrical in shape with a total height of 5.8 m (19') with an inside and outside diameter of 12.7 cm (5") and 50.8 cm (20") respectively. The wall consists of three layers of material: the inner layer made up of high temperature refractory, a middle layer of castable insulating refractory, and a carbon steel shell. The thickness of the high temperature refractory and insulating are 5 cm and 13 cm, respectively. A 0.5. million Btu propane burner is used for start-up purposes.

The Stand Pipe

The stand pipe is used for storage and recirculation of hot solids received from the primary cyclone. It is 2.7 m (9') high and has inside and outside diameters of 22.8 cm (9") and 61 cm (24") respectively. The stand pipe is also refractory lined and it is also equipped with discharge and sampling ports.

Primary Cyclone

The primary cyclone which is 30 cm (12") in diameter, is refractory lined in a similar manner to the reactor and the standpipe.

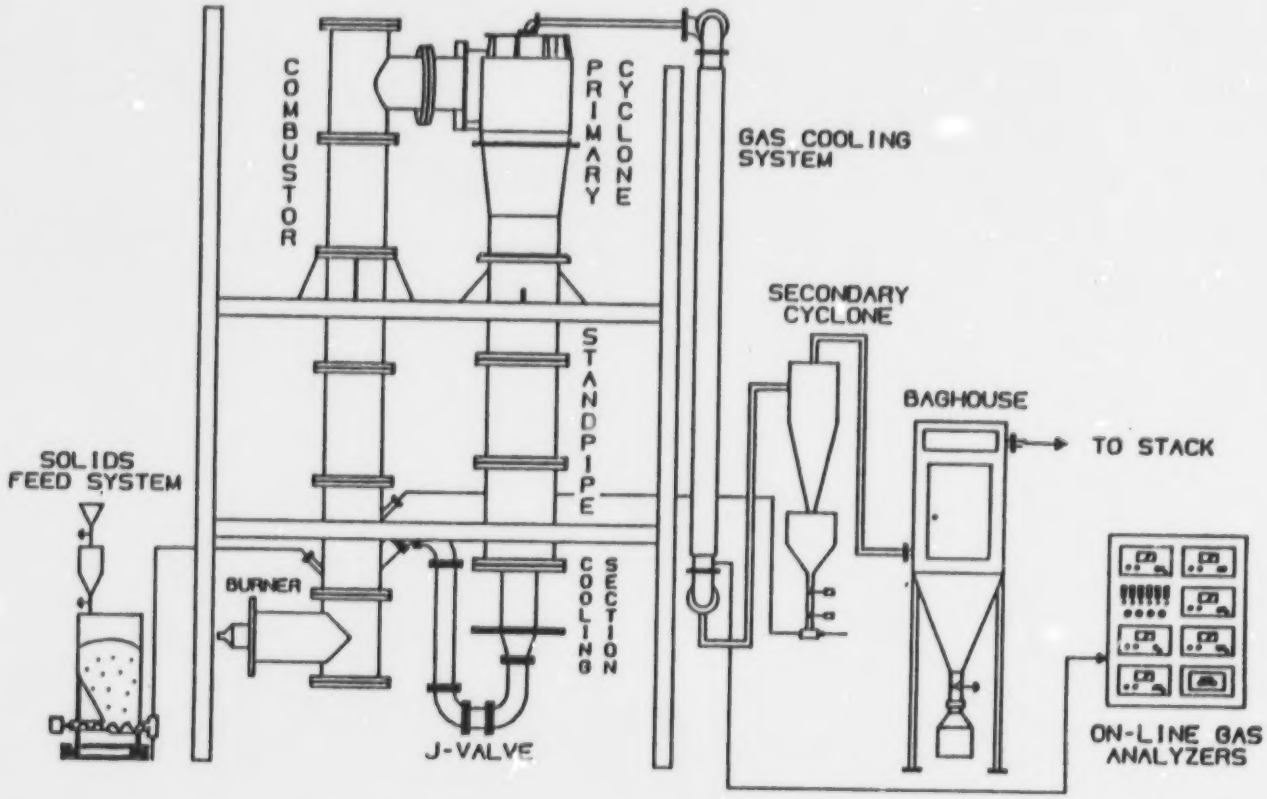

FIGURE 2: RPC CIRCULATING FLUIDIZED BED COMBUSTOR

J-Valve

Circulation of solids from the standpipe to the reactor is achieved by a non-mechanical J-valve. Air is supplied at two points - one at the base of the standpipe, and the other is at the base of the vertical section of the J-valve. The J-valve is made of 2", (Schedule 40) SS pipe. The solids from the J-valve enter the combustor at a location 36.8 cm (14.5") above the distributor plate.

Gas Cooling System

Off-gases from the primary cyclone are fed into a water cooled one pass heat exchanger. It is made of two concentric pipes of nominal diameters of 15 cm (6") (Schedule 40) and 20 cm (8") (Schedule 60), respectively. The height of the exchanger is 5.2 m (17").

The Secondary Cyclone

The secondary cyclone is 25 cm in diameter and made of 304 SS stainless steel. Since the operating temperature is low, the cyclone is neither insulated or refractory lined.

The Baghouse

The flue gases from the cooling section are fed through the secondary cyclone into the baghouse. The latter is a continuous duty self-cleaning dust collector which consist of 16 high temperature bags capable of handling 200 SCFM.

Solid Feeding

Peat was pneumatically fed into the combustor 44.5 cm (17.5") above the distributor plate. The feed rate was regulated using a screw feeder which was placed on a weighing scale to provide an accurate record of the feed rate.

Gas Sampling and Analysis System

On-line gas analyzers, (CO, NO_x, SO_2, CO_2, O_2) are available to monitor the composition of flue gases. The gas sampling position is located after the gas cooling system prior to entering the secondary cyclone. The sampled gas is conveyed to the analyzers through a heated line, a gas moisture conditioning system and a diagphragm pump.

Temperature Measurements

Combustor, standpipe and exit gas temperatures are measured with Inconel sheathed K-type thermocouples, located at approximately 30.5 cm (2') intervals along the height of the combustor and the standpipe.

3.0 PREPARATION AND CHARACTERIZATION OF FEEDSTOCK

The as received thermal peat was screened through 1/8" screen to reject the oversized material. The undersize fraction was fed with no further processing. The as received manganese peat was spread in trays and oven dried to reduce the high mositure content. The moisture was reduced from approximately 80% to 22%. The dried manganese peat was then screened through a 1/8" screen with the undersized fraction being used as the feed. The ultimate and proximate analyses for both peat types as fed to the combustor are shown in Table 1.

Detailed descriptions of the geographical locations of the peats and analyses of different manganese peat samples are given in Appendix B.

TABLE 1

Chemical Analyses of Thermal and Manganese Peats

	Thermal Peat	Manganese Peat
Proximate Analysis (wt%)		
Moisture	40.1	22.2
Ash	1.27	40.52
Volatiles	60.2	35.1
Ultimate Analysis (wt%)		
Carbon - Organic	30.30	20.46
- Inorganic		1.83
Hydrogen	3.33	2.86
Nitrogen	0.51	0.82
Sulphur	0.29	0.55
Oxygen (by difference)	24.2	10.8
Ash Analysis (wt%)		
Manganese	-	16.49
Calorific Value (Btu/lb)	5245	2826

4.0 EXPERIMENTAL PROCEDURE

The reactor standpipe was filled with approximately 200 kg of silica sand. A 0.5 million Btu/hr propane burner was ignited to preheat the recycling sand and the reactor refractory to a temperature of approximately 620°C. This temperature was reached after about 16 hours. The burner was then shut off and coal was fed into the reactor. The heat released by the combustion of the coal raised the reactor temperature to the normal operating temperatures of 850°C in approximately 5 hours. Once the entire system had reached 850°C (ie. combustor and standpipe) the coal feed was stopped and peat was fed to the system from another feeder.

The peat and air feed rates were adjusted to obtain the desired oxygen levels in the flue gas. Different excess air levels were required for the combustion of the thermal peat and the manganese peat. The thermal peat was combusted in an oxidizing atmosphere with a flue gas oxygen level of approximately 2.5%. The manganese peat was combusted at two levels of reducing conditions for the purposes of producing an ash from which Mn could be extracted. For the first test the manganese peat feedrate was adjusted to obtain approximately 1.0% O_2 in the flue gas. In the second test the feedrate was increased until the O_2 level in the flue gas was approximately zero.

When the operating conditions had stabilized the test was started. Variables recorded on the hour included air flow rates, peat feeder weight, ash discharges and temperatures in the reactor. The flue gas composition was monitored continuously and the hourly average was recorded.

In order to refill with feed during a test it was necessary to stop the peat feeder and isolate it from the system. The feeder was stopped every four hours for approximately 5 minutes and refilled with peat.

5.0 OPERATING EXPERIENCE

Operation factors of importance are described below:

The first 48 hours of operation with the thermal peat were used to determine the best method to heat the reactor to operating temperatures and to maintain steady operating conditions. The 620°C reached by heating with the propane burner was not sufficient to sustain combustion of the peat. The feed rates of peat necessary to generate excess heat to raise the reactor temperature were too high for the feeding arranged and the system was very slow to heat up. Since coal feed rates were much lower and the system temperature responded quickly; it was easier to feed coal initially to reach combustor temperatures of 850°C before feeding peat.

To maintain steady operation with the thermal peat it was necessary to feed sand into the combustor to replace bed material lost due to attrition. Attrition is a problem when feeding thermal peat because the ash content is very low (1.27%) and the ash fed to the system is less than the losses. Without feeding sand the bed was steadily depleted by approximately 1.0 kg/hr. Eventually the bed height in the standpipe was reduced to the point that material could no longer be recycled by the J-valve. Sufficient sand was fed into the combustor so that a bed overflow from the standpipe was obtained. This insured that the inventory of solids in the system remained constant.

Bed loss was not encountered when feeding the manganese peat since it has a high ash content (40%) and the ash entering the combustor exceeds the losses by attrition. Large bed overflows from the standpipe were obtained when feeding the manganese peat.

The peat feeder operated reliably when feeding the thermal peat. The system pressure where the feed enters is approximately 1 psi, and it is necessary to seal and pressurize the feeder to prevent gas leakage when feeding the solids. The feeder is depressurized when refilling, therefore it must be isolated from

the system. While the feeder was being refilled (5 minutes) the recycle was stopped to conserve heat and the primary air flow was continued to prevent weepage of material through the distributor plate. The hopper held sufficient material to operate for four hours at the feed rate used in this test.

The same feeder was used with the manganese peat. The density of this peat was much greater than the thermal peat consequently the full feeder could operate continuously for a much longer period of time (12 hours).

At the beginning of the first test with manganese peat the feeder blocked frequently in the pneumatic transport line. The feeder was emptied and it was found that the large mixing screw at the bottom of the hopper was compressing the peat into a hard lump. Occasionally a piece of this lump would break off and be fed into the feed screw and result in the blockage of the pneumatic transport line. The feeder was cleaned and there was no further problems for the remainder of the test.

6.0 RESULTS AND DISCUSSION

One combustion test with thermal peat of 68 hours steady state operation was successfully performed. Appendix C contains the hourly record of measured variables such as solid feed rates, discharge rates and flue gas composition. A 6 and 4 hour duration test was carried out with the manganese peat. The experimental conditions and results are summarized in Table 2. The performance of the process is evaluated below, in terms of the mass and energy balances summarized in Tables 3 and 4. Sample calculations are provided in Appendix D.

6.1 Overall Mass Balance

All input and output streams, flowrates and compositions were measured except for the total flue gas flowrate. This was determined by closing the overall mass balance. The dry flue gas composition was determined using continuous gas analysis. From the dry flue gas analysis the flowrate was found using nitrogen as a tie component. Finally the moisture in the flue gas was assumed to equal the difference between the total flue gas and dry flue gas flowrates.

6.2 Carbon Balance

The carbon balance error for the thermal peat test and the first manganese peat test were very low. This indicates that the flue gas and ash analyses and flow rates were accurately determined.

The carbon balance for the second test with manganese peat was less accurate. A source of error in this balance could be the determination of the carbon content in the flue gas. This test was operated under highly reducing conditions (ie. excess air of -35.6%) therefore it is likely that pyrolysis reactions were occurring and some methane and ethane were present in the flue gas. Since the on-line gas analysis did not include hydrocarbon species the amount of carbon measured in the flue gas may have been lower than that present.

TABLE 2
Summary of Experimental
Conditions and Results

Run #	Thermal Peat	Manganese Peat Test 1	Manganese Peat Test 2
Run Duration (hr)	68	5.83	3.8
Feed Rates (kg/hr)			
Peat	26.80	39.40	51.84
Sand	2.32	-	-
Air	108.6	102.4	102.8
Moisture in Peat	10.75	8.75	11.51
Air Distribution (%)			
Primary	56.4	63.9	58.8
Aeration	6.3	6.6	6.6
Lift	4.3	4.3	4.3
Peat Feeder	26.3	25.2	30.3
Sand Feeder	6.7	-	-
Ash Discharge (kg/hr)			
Standpipe	1.29	5.78	7.63
Baghouse	0.22	1.44	4.95
Secondary Cyclone	0.91	9.46	10.29
Temperatures (^{o}C)			
Combustor Bottom	788	804	788
Combustor Top	859	839	824
Standpipe Bottom	848	827	806
Flue Gas Composition			
O_2%	2.5	1.37	0.23
CO_2%	17.1	18.9	19.1
CO%	0.3	0.5	4.2
NO_x ppm	136	-	-
Excess Air %	13.5	-15.6	-35.6
Carbon Conversion %	97.8	93.7	94.7
Wet Flue Gas Flow Rate (kg/hr) (by difference)	135.3	125.1	131.8
Dry Flue Gas Flow Rate (kg/hr) (nitrogen balance)	114.7	110.6	115.1
Moisture in Flue Gas (kg/hr)	20.57	14.66	16.67

TABLE 3

Elemental Mass Balances

Run	Oxygen			Carbon			Ash			Manganese		
	IN	OUT	%ERROR*	IN	OUT	%ERROR*	IN	OUT	%ERROR*	IN	OUT	%ERROR*
Thermal Peat	41.33	41.78	-1.07	8.120	7.949	2.11	2.66	2.20	17.3	-	-	-
Manganese Peat Test 1	35.88	36.38	-1.38	8.782	8.819	-0.42	15.96	16.15	-1.19	2.632	2.565	2.56
Manganese Peat Test 2	39.77	40.24	-1.18	11.56	10.99	4.86	21.01	22.26	-5.95	3.463	3.324	4.03

* $\text{ERROR} = \frac{\text{IN-OUT}}{\text{IN}} \times 100\%$

6.3 Oxygen Balance

The oxygen balance error for all three runs was consistent and very low; the average error was equal to -1.21%. Indicating that the gas analysis was accurate and the method used to determine the moisture in the flue gas was reasonable.

6.4 Ash Balance

The ash balanced reasonably well for the two manganese tests considering the short duration of both tests (-1.19% and 5.95%). The ash balances error for the thermal peat test however was fairly large (17.3%). This is to be expected since the thermal peat has a very low ash content and a slight absolute error results in a large percentage error. The thermal peat ash balance therefore is much more sensititve to variations in the feed composition as compared to the manganese peat.

6.5 Carbon Conversion Efficiency

The carbon conversion efficiency was determined from the analysis of the feed and ash discharge streams according to the following formula:

$$\text{Carbon Conversion} = 1\left(-\frac{\text{total carbon in ash discharge}}{\text{total carbon in feed}}\right) \times 100\%$$

The efficiency for the three tests are shown in Table 2. High conversions were obtained with both peat types. This is to be expected with highly reactive fuels in oxidizing conditions such as in the test with thermal peat (conversion 97.8%). The tests with the manganese peat were carried out in reducing conditions which should not result in efficient carbon conversion because less oxygen was supplied than required to combust the peat completely. With conversion efficiencies of approximately 94% this was not the case. In the second test with manganese

peat the carbon conversion efficiency remained high because of the large increase in CO emissions. Therefore the carbon must have been converted by pyrolysis and/or gasification type reactions. The water balance shows that no net decomposition of water occurred, so gasification was not the likely route to carbon conversion. More probably, the Mn peat fuel which is highly volatile was converted by pyrolysis reactions which involve the decomposition of the solid into lighter gaseous products.

6.6 Energy Balance

The energy balance was calculated based on the heat content of the inputs (peat and air) and the ouputs (sensible and heating value of dry product gas and solids, and heat content of the steam). The reference temperature used was 21°C which was the temperature of the air fed to the reactor. Heat losses were considered to be the difference between the input and output streams. The results are presented in Table 4.

The average calculated heat loss in the system was 25.6% which is not unreasonably due to the large surface area to volume ratio characteristic of all small reactors. This is especially true for circulating fluidized bed reactor systems which have a large standpipe containing inert solids.

In a large scale unit the surface area to volume ratio would be decreased considerably since the volume (thermal output) is proportional to the square of the diameter while the surface area (heat loss) is proportional to the diameter.

TABLE 4

Energy Balances

Run	Thermal Peat	Manganese Peat Test 1	Manganese Peat Test 2
INPUTS (KJ/hr)x10^{-3}			
Peat	327.0	258.7	340.8
Air	-	-	-
Total In	327.0	258.7	340.8
OUTPUTS (KJ/hr)x10^{-3}			
Combustion Gas (dry)			
Sensible Heat	106.8	100.4	102.5
Calorific Value	3.16	5.04	44.4
Water	86.9	61.2	69.0
Baghouse Discharge	1.37	2.32	9.68
Secondary Cyclone Discharge	5.36	20.4	17.7
Standpipe Discharge	1.09	8.41	11.0
Heat Exchanger	31.72	-	-
Total Out	236.4	197.8	254.3
% Losses	27.7	23.6	25.4

7.0 CONCLUSIONS

7.1 Thermal Peat

The thermal peat was successsfully burnt in the CFBC without any operation problems. The only difference in operation as compared to other fuels such as coal and oil shale was the need to feed additional inert material (sand) in order to maintain the bed inventory. This was necessary due to the very low ash content ($\simeq$1%) of the thermal peat.

The carbon conversion efficiency for the thermal peat was very high, 97.8% at an excess air level of 13.5%. This is to be expected from a fuel high in volatile content. However this fuel does produce higher levels of the undesirable product, CO, as compared to a coal fuel at similar excess air level. The CO in the flue gas was 3000 ppm which is six times the levels obtained when combusting Minto coals.

A temperature gradient in the combustor was obtained when burning thermal peat. The temperature at the level where the feed is injected was 70°C lower than the top of the combustor. This is not typical of other fuels which when burnt resulted in a more uniform temperature profile in the combustor. The most likely reason is the high moisture content in the peat (40%) caused a cooling effect where the fuel is injected; the water absorbed large amounts of heat when it vaporized. Fuels previously tested were very low in moisture content (<1%) and did not require significant amounts of heat to vaporize the water.

7.2 Manganese Peat

The manganese peat was successfully burnt in the CFBC. The carbon conversion efficiencies for both test were high at 93.7% and 94.7%. This is very good since the fuel was combusted under reducing conditions. It was necessary to operate the combustor at reducing conditions to keep the manganese in the monovalent

form. The level of CO was very high (4.29% CO) for the test at -35.6% excess air but this is to be expected since there is insufficient oxygen to further combust the CO to CO_2. It is preferable to minimize the CO formation and future processing of the ash will determine whether such severe reducing conditions, as in test 2, are necessary for good manganese recovery.

The manganese peat unlike the thermal peat has a high ash content. Therefore, there was no need to feed additional inert material in order to maintain bed inventory. In fact, large bed overflows were obtained as when burning other high ash fuels such as oil shale.

The manganese content in the product ash was approximately 18% which is a 3 fold increase in concentration as compared to the manganese peat feed. Therefore burning the peat appears to be a viable method of concentrating the manganese in the solids for future extraction.

8.0 RECOMMENDATIONS

The thermal peat was combusted without any operational problems and at high combustion efficiencies. However, because of the low ash there was a need to feed sand whose cost must be included in determining the economics of the process. Since peat is a highly reactive and easily combusted fuel, other methods of combustion which possibly would not require additional inert material should be investigated.

The manganese peat was dried in ovens from a moisture content of over 80% to about 22%. This would contribute significantly to the cost of processing the manganese peat. It is recommended that the maximum moisture content in the manganese peat be determined that can be successfully combusted. Alternative methods such as drying of the feed using flue gas as well as other techniques should also be investigated.

Ash high in manganese content has been produced. It is recommended that this ash should be further processed in order to determine the best method for leaching the manganese metal from the ash.

Opportunities for commercial use of the thermal peat should be identified. In circumstances where a supply is convenient to the user, such as with a fish plant on the NE coast could make a financially viable project.

9.0 REFERENCES

1. Northland Associates Limited, "ENFOR Project P-294 Review of Commercial and Industrial Wood/Peat Energy in Atlantic Canada 1978/83 and Beyond", pp159-168, DSS Contract 09SC.KH205-3-0179.

2. Himmelblau, D.M., "Basic Principles and Calculations in Chemical Engineering" 3rd Edition. Prentice Hall Inc. New Jersey, 1974.

10.0 AUTHORIZATION

DEPARTMENT: APPLIED SCIENCES

PROJECT TITLE: INVESTIGATIONS ON PEAT ENERGY FOR PROCESS HEAT AND MANGANESE PRODUCTION

PROJECT NUMBER: 05000234 PROPOSAL NUMBER

REPORT AS/87/27

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APPENDIX A

Work Statement

Supply and Services Canada | Approvisionnements et Services Canada

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CONTINUATION

CONTRACT SERIAL NUMBER / NUMÉRO DE SÉRIE DU CONTRAT
23216-6-6286/01-SZ

DSS FILE NUMBER - NUMÉRO DU DOSSIER MAS
SISZ 23216-6-6286

STATEMENT OF WORK

You shall carry out the work described in the technical section of your proposal entitled, "Peat Energy for Process Heat and Manganese Production" dated 8 November 1985, forming a part of this contract, and amended as follows:

You shall conduct two (2) test runs with peat fuel in the circulating fluidized bed combustor. Each test run shall be limited to approximately 100 hours furnace time after all modifications to the feed mechanism and instrumentation checks are completed.

Test '1' shall be conducted on thermal milled peat.

Test '2' shall be conducted on manganese peat.

A report describing the test apparatus complete with sample point locations shall accompany the Draft Final Report. All experimental data shall be recorded and included as an appendix to the Draft Final and the Final Report. Complete mass energy balance and carbon/ash loss calculations for each test run shall be made and reported on in both the Draft Final and the Final Report.

APPENDIX B

Details of Geographical Locations of Peats Tested and Analyses of Manganese Peat

THREE-D GEOCONSULTANTS LIMITED

P.O. Box 3133, Station B, Fredericton, N.B. E3A 5G9
Suite 4B, 108 Prospect Street West
Tel: (506) 453-7700

11 November 1986 Our File: 8622

Dr. Philip Salib
Research and Productivity Council
College Hill Road
P. O. Box 6000
Fredericton, N. B.
E3B 5H1

Dear Philip:

Re: Sampling of Manganese 'Bog' Deposits

Enclosed are location sketches for the sample sites at Dawson Settlement, North Renous, and Grainfield. A total of ten sites were sampled at Dawson Settlement. Recent logging access roads and single party ownership suggest this deposit to be the easiest for acquisition of a bulk sample if analysis results are acceptable.

Six sites were sampled at North Renous. Portions of the deposit were extracted about 1968 and the most accessible material has been removed. To obtain material from this deposit now, brush clearing from abandoned roads would probably be required. Steep slopes would preclude taking trucks close to the deposit and material would have to be transported considerable distances in the bucket of the loader. It is likely that more than one landowner would need to be contacted.

Three sites were sampled at Grainfield. The deposit has good access characteristics but is relatively small and thin. Close proximity (±100 feet) to inhabited dwellings suggests that greater emphasis would need to be placed on 'reclamation' of the site.

If the analysis results show the deposits to be more or less equivalent, the Dawson Settlement site would be the preferred location for bulk sample acquisition.

Please contact me if there are any questions.

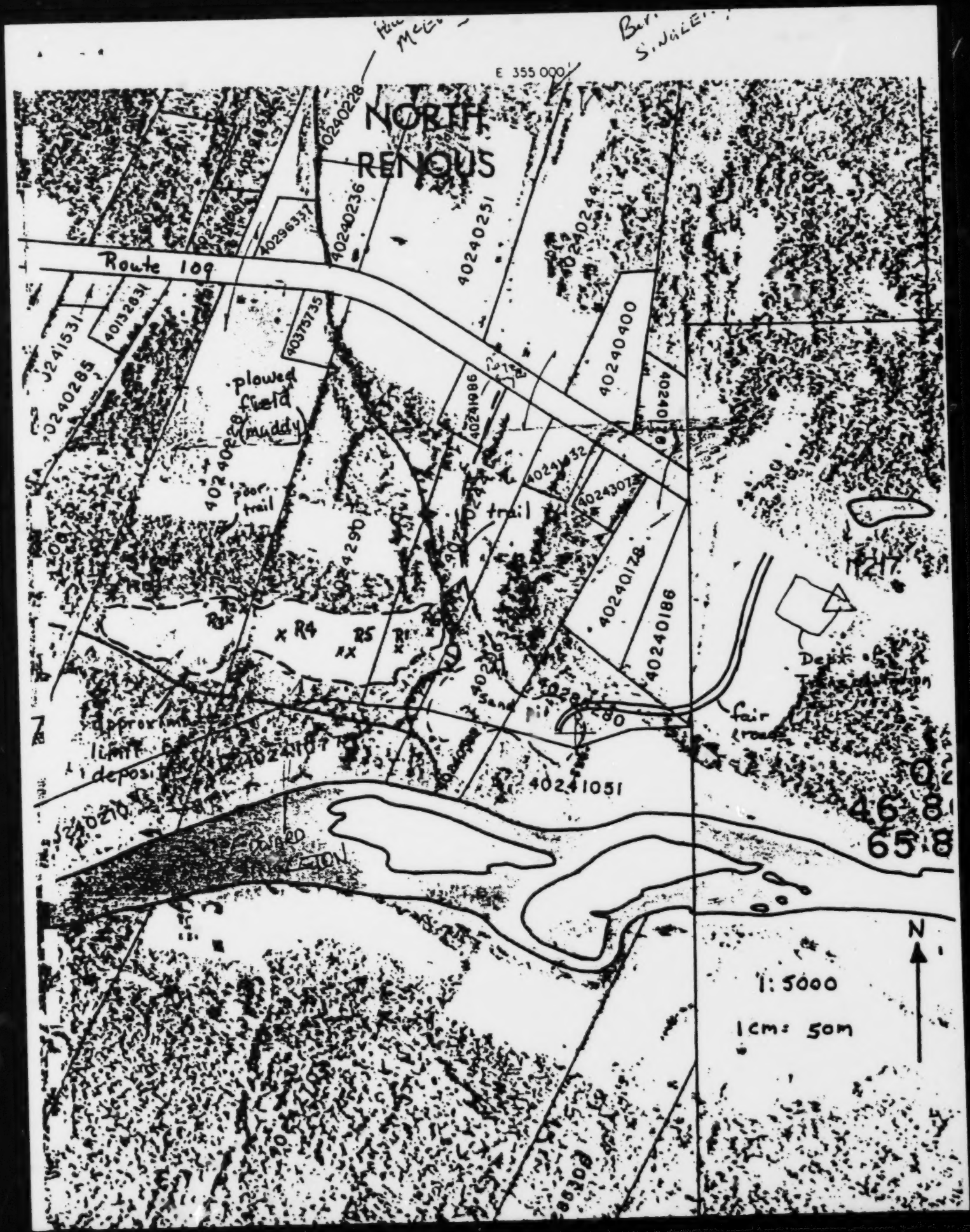
Yours truly,

THREE-D GEOCONSULTANTS LIMITED

David Keys

David Keys
Senior Geologist

DK:bc
Encls.


E 355 000'
NORTH RENOUS
Route 109
plowed field (muddy)
poor trail
trail
fair road
Dept. of Transportation
approximate limit of deposit
Sand Pit
R1
R2
R3
R4
R5
1:5000
1cm = 50m
N

GRAINFIELD
E 353 000
N
1:5000
1 cm: 50m
N 339 000
Road
Bridge
Road

Fo
3424
3421
617662
5000047
5042171
816920
DAWSON SE
1:1000
1 cm = 10m
approximate position of road 617035
709634
816912
816557

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22 December 1986 Our File: 8622

Dr. Philip Salib
Research and Productivity Council
College Hill Road
P. O. Box 6000
Fredericton, N. B.
E3B 5H1

RECEIVED
DEC 2
RPC

Dear Philip:

Re: Bulk Samples of Thermal and Manganese Peat

As per our ongoing discussions, the above two bulk samples have now been delivered to RPC's Fredericton facility.

The fuel (or thermal) peat originated from the so-called Inkerman Ferry deposit (No. 529) in the Shippagan area of northeast New Brunswick. Peat from this deposit has periodically been used for testing of combustion characteristics. It was the designated supply for the peat-fired boiler of Lameque. Three-D has previously arranged shipments from this deposit to Vancouver for use in an NRC-sponsored test of a hog-fuel burner.

Authorization to proceed with the acquisition of this peat was received after the normal peat harvesting season had ended. The material obtained was the stockpiled remnant of a large shipment of fuel grade peat to Florida. We were very fortunate to be able to obtain this material. The alternative would have been to wait until the summer of 1987. The delivered material consists of 203 bags each of which have a volume of four cubic feet and which weigh approximately 80 pounds. The peat is composed primarily of Sphagnum with minor amounts of other peat constituents and has a H4 to H6 degree of humification (decomposition) on the von Post scale. The material has an anticipated gross calorific value (dry basis) of 8000 to 9000 BTU per pound but the 50 to 60% moisture content (wet weight basis) considerably reduces the effective calorific value. The peat was screened prior to packaging to remove larger twigs and other foreign matter.

With regard to the manganese peat, the literature search indicated the Dawson Settlement deposit to have suitable characteristics but initial field sampling suggested the quality was marginal for the proposed use. Additional samples from other areas of this deposit plus from the North Renous and Grainfield deposits

... p. 2

were collected. (See earlier report on sampling locations.) Based on the results of this program, a site on the North Renous deposit was selected for the excavation of a truckload of material. The location of the excavation was the area indicated as R6 on the accompanying map. Access to the site was via a trail running south from Route 109 through the properties of Edward and Bernard Singleton. About 100 metres of trail were 'constructed' through sand pits and alder-covered areas from the end of the road to the sampling location. The 30 centimetre snow cover at the time of sampling in combination with the steep hill on the access trail resulted in the need to transport the material from the site to Route 109 using a loader. The truck was loaded at the roadside. The wet conditions at the site in combination with brush cover and stump content precluded the use of the loader for the actual excavation. A backhoe was obtained and used to excavate the material and load the bucket of the loader. This allowed selective excavation and avoided many of the larger tree pieces within the deposit. As well surface vegetation was easily stripped and contamination by underlying materials was minimized.

As you are aware, the area of the excavation is included in a large parcel of land which was removed from staking under Order-in-Council 84-774. Peat is generally considered a quarriable substance and as such is the property of the surface landowner. In view of these factors and your instructions, no efforts were made to acquire mineral rights.

We trust the material provided to you meets with your satisfaction. Please contact us if there are any questions.

Yours truly,

THREE-D GEOCONSULTANTS LIMITED

David Keys

David Keys
Senior Geologist

DK:bc
Encl.

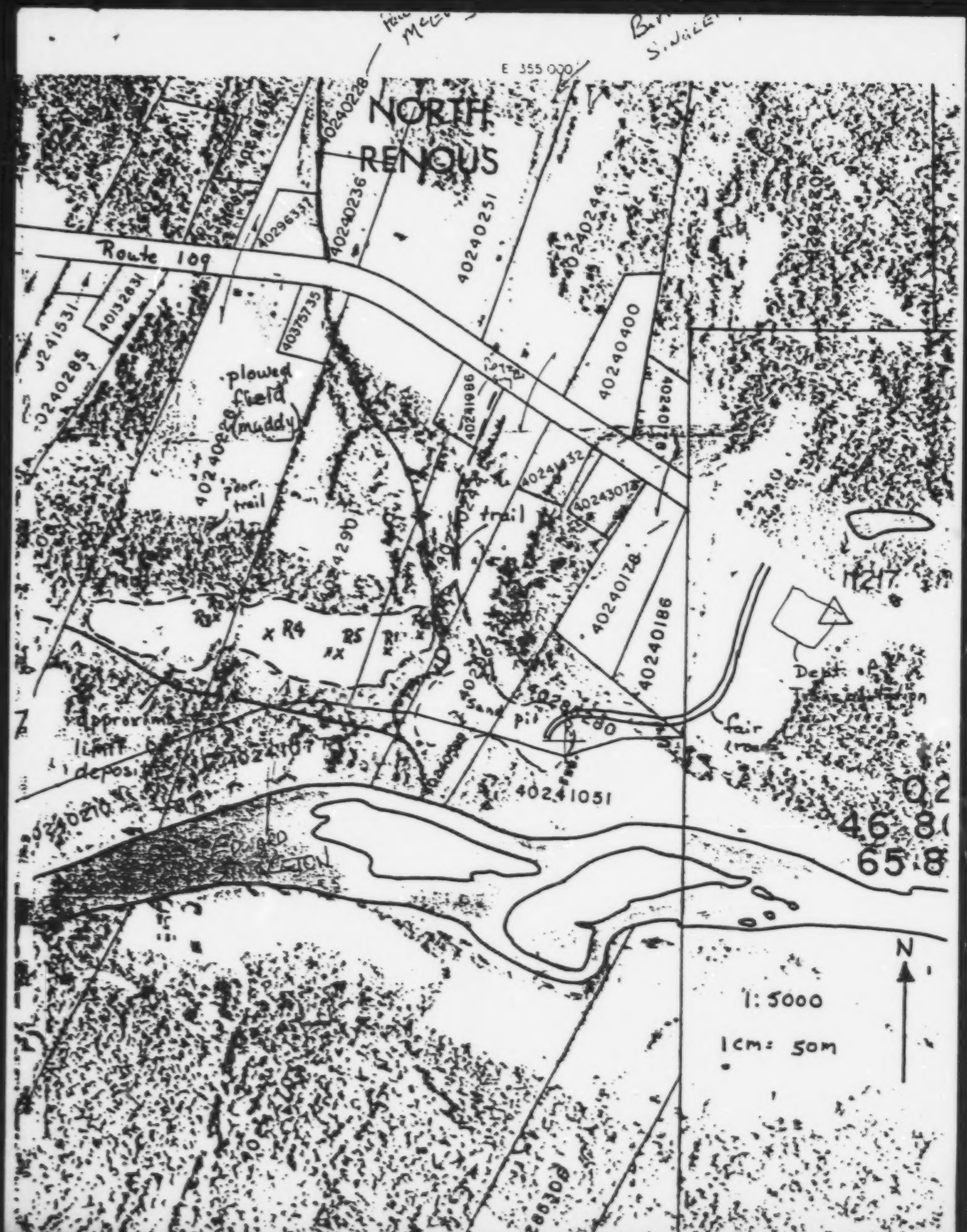
E 355 000
NORTH RENOUS
Route 109
plowed field (muddy)
poor trail
trail
Sand Pit
fair trail
Dept. of Transportation
approx. limit of deposit
1:5000
1cm = 50m
N

Table 5

Analysis of Manganese Peats

Sample	As Received Moisture	Ash	Mn(%)	C(%)	Dry Basis Ash	% Mn	% C
52R-1	80.4	3.20	0.60	8.45	16.32	3.05	43.1
53R-2	85.3	5.42	0.63	3.10	36.88	4.30	21.2
54R-3	79.4	13.4	4.55	2.78	64.86	22.1	13.5
55R-4	68.2	24.1	10.1	1.45	75.80	31.9	4.56
56R-5	76.7	4.06	0.05	9.58	17.43	0.21	41.1
57R-6	83.4	4.58	1.24	6.52	27.57	7.46	39.3
58R-7	45.2	46.4	11.9	1.65	84.68	21.70	3.01
59R-8	74.6	15.2	6.75	4.17	59.66	26.59	16.4
60R-9	82.5	7.14	2.59	4.71	40.79	14.82	26.9
6110-6	68.5	26.1	9.18	0.56	82.72	29.14	1.79
6210-7	67.3	25.8	3.90	1.71	78.83	11.92	5.22
6310-8	42.7	49.6	1.33	2.46	86.54	2.32	4.29
6410-9	51.6	40.1	10.99	1.29	82.92	22.70	2.67
6510-10	62.2	28.6	10.63	2.33	75.72	28.13	6.16

F. E. Campbell
Laboratory Supervisor

FEC/mb/ICAS6

APPENDIX C

Hourly Recordings of Combustor Condition for the Thermal Peat Test

TABLE 6

Combustor Experimental Conditions and Results

time (hr)	feed rate (kg)	sand (kg)	sec.cyc dis (kg)	baghse dis (kg)	stnp dis (kg)
1	29.8	8.2	1.7		
2	30.6	6.9	1.2		
3	29.2	6.7	1.3	0.9	0.2
4	27.0	5.7	1.1		
5	21.4	4.8	0.9		0.7
6	26.6	5.8	0.8	0.8	0.5
7	26.4	5.7	0.8		0.4
8	26.4	6.0	0.8		0.3
9	23.0	5.9	0.9	0.7	11.7
10	25.4	3.7	0.7		0.8
11	27.2	3.2	0.8		0.4
12	27.6	2.8	0.9	0.7	2.1
13	25.4	3.1	0.8		2.8
14	25.2	3.1			
15	30.8	0.0	2.0	0.7	9.5
16	26.4				
17	26.2		1.7	0.4	0.7
18	26.4				
19	23.6		2.4	0.5	0.4
20	23.8	6.6			
21	28.4	6.7	2.6	0.5	0.5
22	30.4	6.5	1.3		1.7
23	28.0	4.9	0.9		0.6
24	29.4	3.7	1.0	0.7	0.5
25		4.8	1.3		4.1
26	29.4	5	1.1	0.6	0.5
27	28.8	5.0	1.0		0.3
28	28.6	2.5	1.0	0.6	0.7
29	23.2	4.3	1.0		2.7
30	28.0	5.2	0.9		3.1
31	27.8	5.1	0.9		3.1
32	27.6	1.4	1.1	1.0	0.9
33	25.0	2.5	1.2		0.9
34	28.0	2.5	0.9		3.9
35	27.6	2.5	0.8	0.7	0.2
36	26.6	0.0	0.9		0.3
37	24.0	2.6	0.8		0.3
38	26.2	0.9			

Combustor Experimental Conditions and Results

time (hr)	feed rate (kg)	sand (kg)	sec.cyc dis (kg)	baghse dis (kg)	stnp dis (kg)
39	25.8	4.4	1.6	0.8	1.8
40	26.6	0.2			
41	24.8	0.2	1.6	0.4	0.7
42	25.6	1.8			
43	27.4	1.0	1.5	0.4	0.5
44	26.0	2.1			
45	26.0	0.9	1.7	0.4	4.6
46	28.2	0.9	0.8		
47	28.2	4.2	0.8	0.5	1.5
48	29.4	3.6	0.9		4.5
49	23.4	1.2	0.8	0.5	0.2
50	29.2	1.9	0.8		2.8
51	30.8	2.7	1.0	0.5	1.7
52	30.2	1.3	0.9		1.5
53	23.0	3.9	0.8	0.6	3.8
54	25.4	2.1	0.8		0.2
55	26.6	2.0	0.7		0.2
56	28.2	0.0	0.8	0.5	1.1
57	24.2	2.0	0.7		5.2
58	25.4	0.0	0.8		0.2
59	25.6	2.0	0.8	0.5	1.6
60	26.8	2.0	0.8		0.3
61	26.2	0.0	0.8	0.3	0.2
62	25.6	2.0			
63	30.4	2.0	1.8		1.1
64	27.2				
65	24.2	2.0	1.6	0.8	0.6
66	28.6				
67	23.2	2.0	1.8	0.4	0.8
68	28.2		0.8	0.2	0.3
average	26.80	2.32	0.91	0.22	1.29
stand. dev.	2.19				

Combustor Experimental Conditions and Results

time (hr)	temp bot (°C)	temp top (°C)	stnp mid (°C)	stnp bot (°C)
1	749	831	822	820
2	773	858	838	846
3	776	856	848	852
4	779	862	848	847
5	773	851	832	835
6	797	853	842	843
7	790	863	843	848
8	790	868	847	854
9	780	851	835	842
10	783	853	837	841
11	783	859	840	845
12	765	864	825	854
13	781	857	844	849
14	750	842	800	841
15	800	867	854	858
16	777	867	844	852
17	809	866	844	852
18	766	864	837	846
19				
20	760	844	830	841
21	757	851	836	839
22		864	846	853
23		858	840	847
24		868	851	857
25		837	812	825
26		865	846	853
27		865	849	853
28		866	855	859
29		857	843	850
30		871	856	857
31	790	871	858	862
32	781	861	849	854
33	800	861	843	849
34	797	869	847	855
35				
36	792	870	859	856
37	807		851	852
38	815	872	854	858

Combustor Experimental Conditions and Results

time (hr)	temp bot (°C)	temp top (°C)	stnp mid (°C)	
39	789	870	848	855
40	803	868	849	857
41	771	846	823	843
42	807	868	853	858
43	808	875	857	864
44	807	850	852	860
45	795	872	852	857
46	812	864	850	854
47	784	862	845	849
48	793	860	847	849
49	781	857	838	842
50	790	858	844	850
51	784	864	850	852
52	786	862	852	858
53	791	849	835	836
54	790	838	820	829
55				
56	790	859	843	846
57	789		831	832
58	780	833	822	830
59	793	839	827	830
60	775	848	831	834
61	800	846	833	837
62	797	855	836	843
63	800	868	852	860
64	808	870	858	862
65	799	860	850	853
66	815	870	859	863
67	779	840	825	830
68	786	862	845	851
average	788	859	842	848
stand. dev.	15	11	12	10

Combustor Experimental Conditions and Results

time (hr)	CO2 (%)	O2 (%)	CO (%)	NOx (ppm)	N2 (%)
1	18.0	1.5	0.8		79.6864
2	18.2	1.3	0-0.9		80.0364
3	18.2	1.0	0-1.0		80.2864
4	17.5	2.1	0-0.5		80.1364
5	18.2	1.8	0.4	160	79.7840
6					
7	17.6	2.1	1.0	130	79.7870
8	17.0	2.7	1.0	160	79.7840
9	16.4	3.5	0.1	140	79.9860
10	16.6	3.6	0.1	170	79.7330
11	16.7	2.9	0.1	150	80.2850
12	16.6	2.9	0.3	120	80.1880
13	17.3	2.6	0.1	138	79.9862
14	16.4	2.7	0.1	125	80.7875
15	17.6	2.3	0.8	128	79.2872
16	17.4	2.0	1.0	118	79.5882
17	17.7	1.3	0.5-1.0	130	80.2370
18	17.8	1.0	.5-1.5	153	80.1847
19					
20	17.8	1.0	0.5-1.0	140	80.4360
21	17.6	1.3	1.0	136	80.0864
22	17.6	2.1	0.4		79.8864
23	17.0	3.0	0.1		79.8864
24	17.0	2.5	0.4		80.0864
25	17.4	2.1	0.0-1.1		79.9364
26	16.6	1.8	0.0-0.6		81.2864
27	17.0	2.5	0.0-0.5		80.2364
28	16.8	2.7	0.0-0.3		80.3364
29	16.0	3.2	0.0-0.2		80.6864
30	16.9	2.6	0.0-0.4	152	80.2848
31	17.0	2.6	0.0-0.4	142	80.1858
32	17.0	2.7	0.0-0.2	143	80.1857
33	16.6	3.1	0.0-0.1	145	80.2355
34	16.6	2.9	0.0-0.4	138	80.2862
35	16.6	2.8	0.0-0.2	124	80.4876
36	15.9	3.1	0.0-0.1	139	80.9361
37	16.9	2.7	0.0-0.2	93	80.2907
38	17.3	2.5	0.5-1.0	119	79.4381

Combustor Experimental Conditions and Results

time (hr)	CO2 (%)	O2 (%)	CO (ppm)	NOx (ppm)	N2 (%)
39	17.8	1.7	0.0-0.5	109	80.2391
40	17.3	2.9	0.0-0.5	123	79.5377
41	17.4	2.7	0.0-0.5	117	79.6383
42	17.4	2.6	0.0-0.3	131	79.8169
43	17.0	3.0	0.0-0.2	144	79.8856
44					
45	17.5	2.4	0.0-0.5	141	79.8359
46	16.7	3.2	0.0-0.5	147	79.8353
47	17.9	1.5	0.0-0.8		80.1864
48	17.0	2.8	0.0-0.3	132	80.0168
49	17.4	2.5	0.0-0.4		79.8864
50	17.4	2.3	0.0-0.3	140	80.1160
51	16.8	2.8	0.0-0.5		80.1364
52	17.2	2.6	0.0-0.6		79.8864
53					
54	16.2	3.8	0.0		79.9864
55					
56	16.8	2.7	0.0-0.3		80.3164
57	16.3	3.3	0.0		80.3864
58	16.5	3.3	0.0-0.1		80.1364
59	15.6	4.0	0.0		80.3864
60	16.8	2.7	0.0-0.4		80.2864
61	16.9	2.6	0.0-0.3		80.3164
62	15.2	4.0	0.0-1.0		80.2864
63	17.8	1.5	0.0-0.3		80.5164
64	17.7	1.8	0.0-0.5		80.2364
65	17.4	2.7	0.0-0.2		79.7864
66	17.5	2.6	0.5-1.0		79.1364
67	17.9	1.5	0.5-1.0		79.8364
68	18.0	1.3	0.5-1.0		79.9364
average	17.1	2.5	0.3	136	80.0864
stand. dev.	0.6	0.7	0.3	16	

APPENDIX D

Mass and Energy Balances
Sample Calculations

Mass and Energy Balances
Sample Calculations

Sample Case: Run #1 Thermal Peat - BASIS (1 hour)

Given (a)		Mass Flowrate (kg)	Temperature °C
INPUTS	- Air	- 108.6	21
	- Peat	- 26.80	21
	- Sand	- 2.32	21
OUTPUTS	- Standpipe Discharge	- 1.29	859
	- Baghouse Discharge	- 0.22	859
	- Secondary Cyclone Disharge	- 0.91	859

(b) Solid Streams Composition:

Element (wt%)	Thermal Peat	Standpipe Discharge	Baghouse Discharge	Secondary Cyclone Discharge
C	20.30	0.11	1.64	15.4
H	3.33			
N	0.51			
S	0.29			
O	24.2			
Ash	1.27	99.9	65.0	84.9
Moisture	40.1	-	-	-

(c) Flue Gas Composition:

O_2%	- 2.5
CO_2%	-17.1
CO%	- 0.3
NO_x ppm	-136
N_2%(by difference)	-80.0864

d) MW = molecular weight
mol fr = mole fraction
mass fr = mass fraction

Calculate:

1) Average molecular weight of flue gas -

$MW = 32.00x.025+44.01x.171+28.01x.003+136x10^{-6}x30.00$

$+28.01x.800864$

$= 30.85 \frac{kg}{kg\text{-mole}}$

2) Flow rate of flue gas (wet)

= flowrate (air+peat+sand-standpipe-baghouse-secondary cyclone)

= 108.6+26.80-1.29-0.22-0.91

= 135.3 kg

3) Flowrate of dry flue gas based on a nitrogen balance

N2 IN = N2 in air+N2 in peat

= 108.6x0.767+26.8x.0051

= 83.433 kg

$\text{Dry flue gas} = \frac{\text{moles } N_2 \text{ in x MW of dry gas}}{\text{mol fr } N_2 \text{ in dry flue gas}}$

$= \frac{83.433}{28.01} \times \frac{30.85}{(.800864+136x10^{6}/2)}$

= 114.73 kg

4) Moisture in flue gas

= wet flue gas flowrate - dry flue gas flowrate

= 135.3-114.73

= 20.57 kg

5) Excess air:

Air required for complete combustion

$$= \frac{32\ kgO_2}{12.0\ kgC} \times \text{mass fraction C} + \frac{8\ kgO_2}{1.01\ kgH_2} \times \text{mass fraction H}$$

$$+ \frac{32\ kgO_2}{32.06\ kgS} \times \text{mass fraction S} - \text{mass fraction O} \times \frac{kg\,air}{0.233\ kgO_2}$$

$$= \frac{32}{12.01} \times .303 + \frac{8}{1.01} \times .033 + \frac{32}{32.06} \times .0029 - .242 / .233$$

$$= 3.571\ \frac{kg\ AIR}{PEAT}$$

$$\text{Air Supplied} = \frac{108.6\ kg\ Air}{26.8\ kg\ Peat} = 4.052\ \frac{kg\ Air}{kg\ Peat}$$

$$\text{Percent Excess Air} = \left(\frac{\text{air supplied}}{\text{air required}} - 1\right) \times 100\%$$

$$= \left(\frac{4.052}{3.571} - 1\right) \times 100\%$$

$$= 13.5\%$$

6) Carbon balance:

Carbon in = Mass flowrate peat x mass fr carbon

$$= 26.80 \times 0.303$$
$$= 8.120\ kg$$

Carbon out

$$= \frac{\text{Mass flow dry flue gas}}{\text{MW flue gas}} \text{ mol fr carbon} \times \text{MW cabon}$$

+ mass flow discharges x mass fr carbon

$$= \frac{114.73}{30.85} \times (.171 + .003) \times 12.01 + 1.29 \times 0.11 + 0.22 \times .164 + 0.91 + .154$$

$$= 7.772 + 0.1776$$

$$= 7.949 \text{ kg}$$

$$\text{Carbon balance error} = \frac{\text{Carbon In} - \text{Carbon Out}}{\text{Carbon In}} \times 100\%$$

$$= \frac{8.120 - 7.949}{8.120} \times 100\%$$

$$= 2.11\%$$

7) Oxygen balance

Oxygen in

= Mass flow air x mf 0 + Mass flow peat x mf 0 + mass flow water x mf 0

$$= 108.6 \times .233 + 26.80 \times .242 + 26.80 \times .401 \times 16/18.02$$

$$= 25.304 + 6.486 + 9.542$$

$$= 41.332 \text{ kg}$$

Oxygen Out

$$= \frac{\text{Mass flow dry flue gas}}{\text{MW of flue gas}} \times \text{mol fr Oxygen} \times \text{MW Oxygen}$$

+ mass flow water out x mass fr Oxygen

$$= \frac{114.73}{30.85} \times (.025 + .171 + .0032/2 + 136 \times 10^{-6}) \times 32 + 20.57 \times 16/18.02$$

= 23.512 + 18.264

= 41.776 kg

Oxygen balance error = $\frac{41.332-41.776}{41.332} \times 100\%$

= 1.07%

8) Ash balance

Ash In = Mass flow peat x mass fr Ash + Mass flow sand

= 26.80 x .0127 +2.32

= 2.66 kg

Ash Out = Mass flow discharges x mass fr Ash

= 0.91 x .849 + 0.22 x 0.65 +1.29x 0.999

= 2.20 kg

Ash balance error = $\frac{2.66-2.20}{2.66} \times 100\%$

9) Carbon conversion efficiency

= 1 - $\frac{\text{mass carbon out in solids}}{\text{mass carbon in total}}$

= 1 - $\frac{1.776}{8.120}$

= 0.978

Energy Balance

Given e) from Himmelblau (1974)

N_2: $Cp = 6.529 + 0.1488 \times 10^{-2} T - 0.0227 \times 10^{-5} T^2$
O_2: $Cp = 6.732 + 0.1505 \times 10^{-2} T - 0.01791 \times 10^{-5} T^2$
CO: $Cp = 6.480 + 0.1566 \times .0^{-2} T - 0.02387 \times 10^{-5} T^2$
CO_2: $Cp = 6.393 + 1.01 \times 10^{-2} T - 0.3405 \times 10^{-5} T^2$
H_2O (g): $Cp = 6.97 + 0.3464 \times 10^{-2} T - 0.0483 \times 10^{-5} T^2$

WHERE T = K
Cp = kcal/kg-mole

f) Ash $Cp = 0.18 + 1.4 \times 10^{-4} T$

WHERE T = ^{o}C
Cp = kcal/kg

g) Calorific Value Peat = 5245 Btu/lb

CALCULATE

10) Heat Input Assuming T Reference is 21°C or 294K

AIR IN is at 21°C $\therefore \Delta H$ AIR = 0.0

ENTHALPY IN PEAT = mass peat x calorific value

$$= 26.80 \text{ kg} \times 5245 \frac{\text{Btu}}{\text{lb}} \times 1.055 \frac{\text{KJ}}{\text{Btu}} \times \frac{2.205 \text{lb}}{1 \text{ kg}}$$

$$= 327.0 \times 10^3 \text{ KJ}$$

$\therefore$ TOTAL HEAT INPUT = 327.0×10^3 KJ

11) Heat Output T Out = 859°C = 1132K

Specific heat of the dry flue gas $C_p = \sum_{i=1}^{n} X_i C_i$

WHERE C_i = specific heat of individual species i.e. N_2
X_i = mole fraction of C_i
C_p = total average specific heat

$\therefore$ for the flue gas $C_p = 6.5107 + 0.2961 \times 10^{-2}\, T - 0.7694 \times 10^{-5} T^2$

Enthalpy of the dry flue gas = $T = \int_{T=294}^{T=1132} CpdT$

$$= 6.5107 \times (1132-294) + \frac{2.96 \times 10^{-2}}{2} \times (1132^2 - 294^2) - \frac{.07694 \times 10^{-5}}{3} \times (1132^3 - 294^3)$$

= 6860 Kcal/kgmol

$\therefore \Delta H$ flue gas = $\frac{114.73}{30.85} \times 6860 \times 4.1868 \frac{KJ}{Kcal}$

= 106.8×10^3 KJ

Also include the heating value of the flue gas because of incomplete combustion of CO.

Heating Value = $\frac{114.73}{30.85} \times .003 \times 283.1 \times 10^3 \frac{KJ}{kg\ mol\ CO}$

= $\therefore \Delta H$ Flue Gas Total = 109.96×10^3 KJ

STEAM OUT

ΔH STEAM = ΔH water from 294K-373K + ΔH vapourization water at 373K
+ ΔH steam from 373K to 1132K

ΔH WATER = 20.57 kg x $\frac{1\ Kcal}{Kg\text{-}K}$ x (373-294) x 4.1868 $\frac{KJ}{kcal}$
T= 294-373

= 6.8×10^3 KJ

Δ H VAPORIZATION = 20.57 kg x 970.3 $\frac{Btu}{lb}$ x 1.055 $\frac{KJ}{Btu}$ x 2.205 $\frac{lb}{Kg}$

$= 46.43 \times 10^3$ KJ

ΔH STEAM $= 6.97 \times (1132-373) +$

$$\frac{0.3464.32^2-373^3)}{2} - \frac{-0.0483 \times 10^{-5}}{3} (1132^3-373^3)$$

T = 373-1132)

$= 7044 \frac{KCal}{Kgmol}$

$\therefore \Delta H$ STEAM $= \frac{20.57}{18.02} \times 7044 \times 4.1868 =$

$= 33.67 \times 10^3$ KJ

$\therefore \Delta H$ STEAM OUT TOTAL $= (6.80+46.63+33.67) \times 10^3$KJ

$= 86.9 \times 10^3$ KJ

SOLIDS OUT

Baghouse Discharge Enthalpy = Sensible Heat + Heating Value

Sensible Heat $= 0.22 \text{kg} \times 0.18 \times (859-21) + 0.7 \times 10^{-4} (859^2-21^2) \times 4.1868$

$= 0.19 \times 10^3$KJ

Heating Value Calculated Assuming Heat From Combustion of Carbon

$= 0.169 \frac{\text{kg carbon}}{\text{kg Ash}} \times \frac{\text{1kg mole carbon}}{\text{12.01 kg carbon}} \times \frac{393.7 \times 10^3 \text{KJ}}{\text{kg mole carbon}}$

$= 5376 \frac{kJ}{kg\ Ash}$

$\therefore$ Heating Value $= 0.22 \times 5376$

$= 1.18 \times 10^{-3}$KJ

$\therefore$ Total Enthalpy of Baghouse Discharge = 1.37×10^3 KJ

Enthalpies of the Standpipe Discharge and Secondary Cyclone Discharge are determined in the same way as the Baghouse Discharge

Heat removed by heat exchanger

Water Flowrate = 2.87 kg/min

Water Temperature Rise = $44^{\circ}C$

$\therefore$ $H = 2.87 \frac{kg}{min} \times \frac{60 min}{1 hr} \times \frac{4.1868 kJ}{kg\ water - {}^{\circ}C} \times 44^{\circ}C$

$= 31.72 \times 10^3$ KJ/hr